Available online at www.sciencedirect.com

ScienceDirect

Energy Procedia 63 (2014) 1151 – 1161

Energy Procedia

GHGT-12

Performance comparison of moving and fluidized bed sorption systems for an energy-efficient solid sorbent-based carbon capture process

Kiwoong Kim[a], Yong-Ki Park[b], Junhyung Park[a], Eunji Jung[a], Hwimin Seo[b], Huiyong Kim[a], Kwang Soon Lee[a,*]

[a]*Department of Chemical and Biomolecular Engineering, Sogang University, Seoul, 121-742, South Korea*
[b]*Green Chemistry Division, Korea Research Institute of Chemical Technology (KRICT), Daejeon, 305-600, South Korea*

Abstract

As a future post-combustion CCS technology, the solid sorbent-based process is extensively researched recently because lower energy penalty is expected thereof compared to that of the conventional amine scrubbing process. The type of gas-solid contactor of solid sorbent-based process has considerable impact on overall performance but systematic comparison of the different designs has not been reported yet. In this work, various sorption systems with different gas-solid contact types including i) moving-, ii) bubbling-, and iii) fast fluidized-bed using absorbents of amine supported sorbent are established. Mathematical models of each sorption system are developed from first principles and compared in terms of economy indicators. From the results, an optimal sorption system is discussed and a research direction of process design is suggested to improve the process economy of the solid sorbent-based carbon capture processes.

Peer-review under responsibility of the Organizing Committee of GHGT-12

Keywords: Carbon capture;Solid sorbent-based process;Moving bed;Fluidized bed

* Corresponding author. Tel.:+82-2-705-8477; fax:+82-2-3272-0319.
E-mail address: kslee@sogang.ac.kr

Peer-review under responsibility of the Organizing Committee of GHGT-12
doi:10.1016/j.egypro.2014.11.125

1. Introduction

Nomenclature

A	cross-sectional area of heat exchange (m^2)
D_{ax}	mass axial dispersion coefficient (m^2/s)
ΔH	heat of absorption (kJ/kmol)
K	mass interchange coefficient between phases in fluidization beds (s^{-1})
N	total number of amine sites (kmol/kg)
P_{CO_2}	CO_2 partial pressure (kPa)
T	temperature (K)
U	overall heat transfer coefficient (kJ/m^2-s-K)
W	required electrical work (kWh/ton-CO_2)
W_d	width of heat exchange plates in CMB (m)
a_{gs}	heat transfer area between the gas and solid phases per unit volume (m)
b	half distance between heat exchange plates in CMB (m)
c_g	specific heat of gas (kJ/kg-K)
$c_{p,s}$	specific heat of solid (kJ/kg-K)
C	concentration in gas phase ($kmol/m^3$)
h_{gs}	heat transfer coefficient between the gas and solid phases (kJ/m^2-s-K)
k	affinity coefficient between CO_2 and the sorbent (1/kPa)
$\dot{m}_s$	solid circulation rate (kg/hr)
$\dot{n}_{CO_2}$	CO_2 reacted in absorber (kmol/hr)
q	absorbed loading of CO_2 in sorbent (kmol/kg)
r	reaction rate with respect to CO2 concentration (kmol/m3-s) in BF and FF beds, mass transfer rate (kmol/kg-s) in CMB
u	gas velocity (m/s)
Greek Symbol	
δ	volume fraction of bubble phase (in BFB), core phase (in FFB) (m^3-bubble or core/m^3 bed)
ε	volume fraction of solid (-)
γ_b	volume fraction of solid in bubble phase in BFB (-)
λ_{ax}	effective axial thermal conductivity of gas (kJ/m-s-K)
ρ_s	solid density (kg/m^3)
Abbreviation	
BFB	bubbling fluidized bed
CMB	counter-current moving bed
FFB	fast fluidized bed
HTW	heat transfer water

Global warming is caused by emissions of huge amount of greenhouse gases, mainly carbon dioxide, from large stationary point sources such as fossil fuel based power plants [1]. In the light of growing population demanding massive amount of energy, it is expected that the usage of fossil fuels will increase continuously in the foreseeable future [2]. Under these circumstances, CCS technology (carbon capture and sequestration) is a viable alternative to mitigate the effects of global warming by reducing atmospheric emissions of carbon dioxide. However, high energy requirement is a crucial problem of current CCS technology with 70-80 % of the total energy consumed during the capture process [3]. Thus, many groups have focused on the development of carbon capture processes that have low energy demands to reduce overall capture cost.

Table 1. An epitomic overview of literature survey on process configurations of solid sorbent-based post-combustion CO_2 capture process using temperature swing.

References	Reactor types		Selected sorbents	Regeneration. temperature	Development status
	Absorber	Desorber			
Charitos et al. [4]	BF	FF	CaO (calcium oxide)	800-900°C	10 kW_{th} plant
Rodríguez et al. [5]	FF	FF	CaO	800-900°C	30 kW_{th} Plant
Modekurti et al. [6]	Staged BF	MB	NETL-32D (amine impregnated sorbent)	<200°C	under dev.
Kim et al. [7]	MB	MB	Zeolite-13X	<170°C	under dev.
Kim et al. [8]	FF	FF	K_2CO_3/Na-promoted MgO/Li_4SiO_4	< 750°C (max)	under dev.
Veneman et al. [9]	FF	BF	SAS (supported amine sorbent)	<150°C	under dev.
Ströhle et al. [10]	FF	FF	CaO	800-900°C	1MW_{th} plant

The CCS technology closest to commercialization is an amine scrubbing process using MEA (monoethanolamine) as the solvent. The MEA scrubbing process uses 3.5-4 GJ per tonne of CO_2, which is mainly consumed for the regeneration of the solvent. This amount of thermal energy equals to more than 20% decrease of electricity output of an existing power plant when MEA scrubbing process is retrofitted to the plant [11,12]. In addition to the high-energy consumption, solvent loss by oxidative and thermal degradation is not negligible, and the environmental problems caused by the use of amines also should be carefully addressed.

One alternative solution to the amine scrubbing process is a solid sorbent-based carbon capture process [13-15]. Solid sorbent-based processes are believed to require low regeneration energy due to lack of water evaporation and the low heat capacity of solids and also environmentally benign. Thus, numerous recent researches can be easily found and Table 1 shows solid sorbent-based carbon captures processes under development. Various types of reactors, sorbents, and different operating temperature are proposed in these studies and it is still uncertain which combination of these factors will lead to economically feasible process.

One of the most ambiguous parts for designing a solid sorbent-based process is selection of the reactor type for the absorber and desorber. The operating temperature can rather easily be decided, either high or low, depending on type of sorbent used. A regenerable CaO (calcium oxide) sorbent can be used in high temperature process and an amine (polyethylenimine, PEI) impregnated sorbent is mostly used in low temperature process. However, choosing an appropriate gas-solid contactor between BF (bubbling fluidized), FF (fast fluidized), and CM (counter-current moving) beds is confusing. The reactor type doesn't seem to directly depend on absorption mechanism nor type of sorbents.

In this work, mathematical models for solid sorbent-based carbon capture processes using BF, FF, and CM beds are established and performances of each reactor type are compared in terms of economic indicators to analyse pros and cons of each configuration.

2. Process description

2.1. Flue gas and amine sorbent

Identical flue gas and amine sorbent are assumed for all configurations for fair comparison. Also, it was assumed that the developed carbon capture processes are retrofitted to a 500 MWe PC power plant with a net electrical efficiency of 44.3 %. The flue gas is constituted by 13.2 mol% CO_2, 9 mol% H_2O, and balanced N_2 with a mass flow rate of 546 kg/s and 300 K.

Table 2. Physical properties of chosen amine sorbent.

Sorbent	Amine (PEI) impregnated sorbent
Specific heat (kJ/kg-K)	1.25
Particle diameter (μm)	200 (BF and FF beds) / 3,400 (CMB)
Heat of absorption (kJ/kmol-sorbent)	50.03
Bulk/particle density (kg/m^3)	1000 / 1800
Amine concentration (kmol/kg)	0.00298

The low temperature solid sorbents of amine (PEI) immobilized on silica supporter is selected as sorbent because the amine sorbents has advantages such as i) its low energy requirements ii) low toxicity iii) high capture capacity iv) low regeneration temperature v) large surface area [16]. The physical properties of amine sorbent were denoted in Table 2. In CM bed, the particle diameter is 3,400 μm that is large enough to avoid fluidization and we assumed that the properties are identical irrespective of particle diameter. The attrition of solid sorbent during movement in CMB may raise a problem, however, in this work we assumed that the amine sorbent is hard enough (high attrition index, AI) to avoid attrition problem.

Effects of H_2O adsorption onto the amine sorbents are not considered seriously even though the amine sorbents is known to react with H_2O and CO_2 simultaneously. It has been reported that CO_2 working capacity is slightly increased in the hydrous condition[17-19] and Hoffman et al. [17] showed that there is little competition for adsorption sites of CO_2 and H_2O. Also, the rate of adsorption of H_2O is an order of magnitude of 10 times slower than that for CO_2 and H_2O adsorption can be avoided via controlling gas-solid contact time.

2.2. Process configuration

Figure 1 represents a generic carbon capture process using temperature swing. The flue gas is fed to an absorber and reacted with a regenerated solid sorbent. The exhausted gas is purged and the reacted solid sorbent is transferred to a regenerator and CO_2 is desorbed from the sorbent by regenerative heat supply. In a liquefaction stage, the desorbed CO_2 is reached to end state of 110 bar and 298 K. In a heat exchanger, the solid sorbents from absorber and from regenerator are heat exchanged before it is supplied to regenerator and absorber, respectively. The adsorbed water content in amine sorbent is depleted in a drying process using the exhausted gas from the absorber.

The absorber and regenerator can have various reactor types such as BF, FF, and CM beds. In this work, only absorber is varied to have different reactor types and regenerator is assumed to operate in gas-solid equilibrium (independent of reactor type) in the light of the fast kinetics of regeneration compared to the kinetics of absorption.

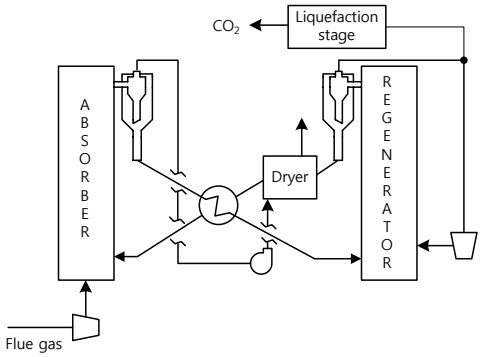

Fig. 1. Conceptual diagram of generic post-combustion carbon capture process using temperature swing. In BF and FF beds, the gas and amine sorbents are fed at adjacent bottom of bed, whereas, in CM bed, the gas is fed at the bottom of bed and the amine sorbent flows down from the top of bed.

3. Modeling of the processes

Modeling details of three different beds are described and model for techno-economic analysis is also explained in this section. In fluidized bed processes, steady state models are developed and in CM bed, dynamic models were constructed because of the solution convergence problem resulted from the nonlinearity of models. An additional assumption is that the regenerative energy of H_2O is negligible through minimizing H_2O adsorption by properly controlling gas-solid contact time.

3.1. Reaction models

An equilibrium model and corresponding model parameters of solid sorbent with CO_2 from the experimental study of Ebner et al. [20] is used. The equilibrium model is found as

$$q_{CO_2} = \frac{kP_{CO_2}N}{1+kP_{CO_2}}, \text{ where } k = \frac{k_f k_H}{k_b} = k_0 \exp\left(-\frac{\Delta H}{RT}\right). \tag{1}$$

ΔH is a sum of difference between activation energy for the forward and backward reactions and heat of adsorption for physi-sorption.

A quasi-chemical kinetics of Bohart-Adams model was used for describing the mass transfer of CO_2 with amine sorbents [21,22].

3.2. Bubbling and fast fluidized beds

The fluidized bed models were developed based on KL model [23]. It is assumed that the temperature inside the bed is uniform and the solid particles are perfectly mixed without solid conversion distribution.

In BFB, two regions (bubble and emulsion phases) are modeled with an interchange coefficient between thereof. Based on the KL model, the axial CO_2 concentration in the bubble and emulsion phases with the bed height, z, is represented as follows

$$\begin{aligned} \delta u_b^* \frac{dC_{b,CO2}}{dz} &= \delta\gamma_b r_{CO_2,b} + K_{be}\delta(C_{b,CO_2} - C_{e,CO_2}), \\ (1-\delta)u_{mf}\frac{dC_{e,CO2}}{dz} &= (1-\delta)(1-\varepsilon_{mf})r_{CO_2,e} - K_{be}\delta(C_{b,CO_2} - C_{e,CO_2}). \end{aligned} \tag{2}$$

In FFB, unlike BFB where volume fractions of bubble and emulsion phases are constant, the fraction of two regions (core and wall phases) is fixed in a lower dense region, whereas, the volume fraction of wall phase is decreased in an upper lean region. The axial CO_2 concentration in the core and wall phases is as follows

$$\begin{aligned} \delta(z)(1-\varepsilon^*)u_c\frac{dC_{c,CO2}}{dz} &= \delta(z)\varepsilon^* r_{CO_2,c} - K_{cw}\delta(z)(C_{c,CO_2} - C_{w,CO_2}), \\ (1-\delta(z))(1-\varepsilon_w)u_w\frac{dC_{w,CO2}}{dz} &= (1-\delta(z))\varepsilon_w r_{CO_2,w} + K_{cw}\delta(z)(C_{c,CO_2} - C_{w,CO_2}). \end{aligned} \tag{3}$$

The amount of circulating solid is determined through correlations of gas velocity, solid holdup, bed height, and particle diameter. We used same hydrodynamic models reported in our previous work on the modeling of fast fluidized bed [8].

The mass transfer coefficient between CO_2 and amine sorbent reported in [22] is based on BFB condition, thus, in BFB, the same mass transfer coefficient was used. However, in FFB, the gas velocity is much faster than BFB. To reflect this condition, we used 3 times larger mass transfer coefficient.

The amount of CO_2 reacted in absorber is calculated as

$$\dot{n}_{CO_2} = \dot{m}_s(q_{av} - q_{in}(T_{DE}, P_{CO_2,DE})). \tag{4}$$

q_{in} is an equilibrium concentration of amine sorbent fed to absorber, which depends on the regeneration condition. As shown in Fig. 1, a partial regenerated CO_2 gas is re-circulated to regenerator as sweep gas, $P_{CO_2,DE}$ was set to 1 bar.

In the fluidized bed models, the solid has a representative conversion value inside the bed, whereas, the gas has an axial concentration gradient. These differences made the solution procedure being iterative to balance the overall mass balances between gas and solid phases. We used the gas velocity as the iterative variable and changed it until the overall mass balance is satisfied.

3.3. Counter-current moving bed

The gas and solid flow in an opposite direction in CM while the gas-solid reaction takes place. The gas flows upward and the amine sorbent flows downward by gravity. The solid flow rate is controlled in a pneumatic slide valve placed at the top of absorber. Compared to the constant temperature inside the bed for BF and FF beds, the temperature gradient is shaped in CM because both gas and solid flows belong to plug flow in a reverse direction. The material and energy balance equations of CM are described as follows:

$$\frac{\partial C}{\partial t} = -\frac{\partial (u_g C)}{\partial z} - \rho_s \frac{1-\varepsilon}{\varepsilon} \sum_i r_i, \tag{5}$$

$$\frac{\partial y_i}{\partial t} = -u_g \frac{\partial y_i}{\partial z} + \frac{D_{ax}}{C}\left(C\frac{\partial^2 y_i}{\partial z^2} + 2\frac{\partial C}{\partial z}\frac{\partial y_i}{\partial z} + y_i \frac{\partial^2 C}{\partial z^2} \right) + \frac{\rho_s}{C}\frac{1-\varepsilon}{\varepsilon}\left(y_i \sum_{\substack{j=N_2,\\ CO_2}} r_j - r_i \right), \tag{6}$$

$$\frac{\partial T_g}{\partial t} = -u_g \frac{\partial T_g}{\partial z} + \frac{(1-\varepsilon)\rho_s}{\varepsilon \rho_g c_g} \sum_i c_{g,i} r_{in,i} (T_s - T_g) + \frac{\lambda_{ax}}{c_g \rho_g L^2}\frac{\partial^2 T_g}{\partial z^2} + \frac{h_{gs} a_{gs}}{\varepsilon \rho_g c_g}(T_s - T_g), \tag{7}$$

$$\frac{\partial T_s}{\partial t} = u_s \frac{\partial T_s}{\partial z} + \frac{1}{c_s}\sum_i c_{g,i} r_{out,i}(T_g - T_s) + \frac{1}{c_s}\sum_i r_i(-\Delta H_i) + \frac{h_{gs} a_{gs}}{(1-\varepsilon)\rho_s c_s}(T_g - T_s) + \frac{U_{hs}}{2b(1-\varepsilon)\rho_s c_s}(T_h - T_s), \tag{8}$$

$$\frac{\partial T_{htw}}{\partial t} = -u_{htw}\frac{\partial T_{htw}}{\partial z} + \frac{U_{htw-s} W_d (N_m + 1)}{\rho_{htw} c_{htw} A_{htw}}(T_s - T_{htw}). \tag{9}$$

Eqn. (9) denotes the temperature of HTW (heat transfer water) inside the absorber; HTW is maintained in liquid state by regulating the pressure. HTW recovers the absorption energy while it flows upward inside absorber.

In CMB, as depicted in Fig. 2, the solid heat exchanger installed between absorber and regenerator shown in Fig. 1 is not necessary because of the solid temperature gradient inside the bed. The hot sorbent from regenerator and cold sorbent from absorber are cooled down and heat through HTW. HTW gains absorption heat in absorber and pass through heater; in heater, HTW is externally heated up to MTA (=10K, minimum temperature approach) above regeneration temperature and phase change occurs by supplying latent heat through proper manipulation of HTW pressure line. In HR1, the phase change takes place again and the latent heat is used to provide the regeneration energy and the remained sensible heat is used to heat up the amine sorbent fed to regenerator in HR2. The used HTF is re-circulated to CWR.

In CMB, the particle diameter is much larger than that for fluidized bed condition, which implicates that the behavior of mass transfer may be different with the fluidized bed. Although the absorption reaction is slower than the adsorption, the intra-pore size of amine sorbent is larger than the micro-pore of zeolite. Additionally, because of data availability, we identically used the same mass transfer coefficient of CO_2 reacting with zeolite 13X in our previous work [24].

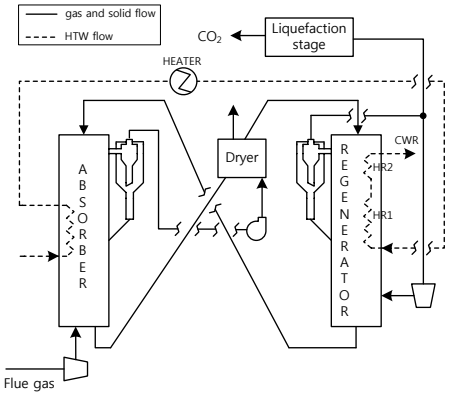

Fig. 2. Conceptual diagram of CMB process.

3.4. Techno-economic analysis model

The BF, FF, and CM beds processes were compared in the aspects of economy analysis which take into accounts operating and capital expenditures.

The operating expenditures comprise of energy used in blowers, compressor, and regeneration. In BF and FF beds, the heat of regeneration is computed using Eqn. (10), which includes sensible and reaction energy. It is assumed that the amine sorbent exchanges their sensible energy with MTA of 10K.

$$\dot{Q}_{reg} = \dot{m}_s c_{p,s}(2\Delta\mathrm{MTA}) - \dot{n}_{CO_2}\Delta H. \tag{10}$$

In CMB, the first term of right hand side of Eqn. (10) is substituted by the energy required to make temperature of HTW higher than the regeneration temperature plus the latent heat. The regeneration energy is converted to the equivalent electrical work using correlation as

$$W_{eq} = 0.75 \times Q_{reg}\left(\frac{T_{DE}+10-313}{T_{DE}}\right). \tag{11}$$

The total work requirement is given by:

$$W_{total} = W_{eq} + W_{comp} + W_{pump}. \tag{12}$$

It was considered that 3 unit processes are built and each unit process takes one third of the flue gas from 500 MWe power plants. After sizing the equipment, PFC (total process facilities capital) was estimated from literature [25] and Aspen Economy Analyzer. The basis of cost data is January 2010. Because the processes in this work are non-conventional, the preliminary capital cost estimate was conducted with restricted cost data. PFC includes equipment and installation costs. In most equipment, MOC (material of construction) was chosen as carbon steel and stainless steel. TCR (total capital requirement) is calculated using the following equation (adapted from Rao et al. [26]):

$$\mathrm{TCR}=\text{owner's cost}+\text{interest}+\mathrm{TPC}(=\mathrm{PFC}+\mathrm{GFC}+\mathrm{EHO}+\mathrm{CPP})+\text{pre-production}. \tag{13}$$

TPC (total plant cost) is the sum of PFC, GFC (general facilities capital), EHO (engineering and home office overhead), and CPP (contingencies – project and process). The owner's costs include royalties, startup costs, inventory capital and pre-production costs are 1 month's cost of fixed and variable O&M cost.

$$\mathrm{TRR}=\mathrm{TCR}\times\mathrm{CRF}+\mathrm{TOM}(=\mathrm{FOM}+\mathrm{VOM}) \tag{14}$$

From Eqn. (14), TRR (total annual revenue requirement) was calculated with CRF (capital recovery factor) of 0.15. COE (cost of electricity) was estimated from the obtained TRR with a plant capacity factor of 0.8 and, consequently, cost of CO_2 avoided was computed.

We hypothetically assumed that the amine sorbent cost is 1.25 $ per kg of sorbent and TRR of the reference power plant is 148.7 M$/year with net output of 457.6 of which data are extracted from Rao et al. [26]. The liquefied CO_2 is transported to 165 km far storage and disposal repository with the cost of $5/ton CO_2.

4. Results

4.1. Sensitivity analysis

The major process operating conditions were relatively changed from -15 to 15 % with a constraint such that the CO_2 recovery is fixed at 85 %. The nominal values of operating parameters are shown in Table 3. The perturbation variables were restricted to satisfy the constraint

Table 3. Nominal process operating parameters that are perturbed in sensitive analysis study.

	Nominal values
BFB	T_{DE} =500 K, u_g =0.15 m/s, H = 3 m
FFB	T_{DE} =500 K, pres. drop =0.15 bar, H = 30 m
CM	T_{DE} =500 K, u_s =0.002 m/s, u_{cw} = 0.2 m/s, H = 2 m

In Fig. 3, the effects of relative variation of selected parameters on the total work are shown. In BFB, the desorption temperature has the most significant effect on the total work. When desorption temperature increases, more fresh sorbent is fed to the absorber so that smaller solid circulation amount is sufficient to achieve the specified CO_2 recovery. However, further increase of desorption temperature negatively acts on the total work because the regeneration energy is increased to maintain such a high desorption temperature, which indicates the existence of an optimal desorption temperature. As gas velocity increases, the volume fraction of bubble phase is increased and this larger bubble phase deteriorates the gas solid reaction, which implies that larger solid circulation is required to maintain 85 % CO_2 recovery. The bed height is reversely affected to the total work; smaller CO_2 is reacted in shorter bed height. In FFB, bed height and pressure drop of absorber have little effect on the total work. As was in BFB, desorption temperature has the similar effect on the total work and it has an optimal temperature at -5 % change at the nominal value.

In CM, the higher desorption temperature represents that HTW has to be heated up to higher temperature, which increases the external energy use and in turn increases the total work. The bed height is insignificant on the total work. As HTW velocity increases, it enhances the gas-solid reaction by removing the heat of absorption, however, more external heat is required to heat HTW up to desorption temperature, which counteracts the total work so that total work is increased. More solid is circulated for the faster solid velocity, which promotes the gas-solid reaction. The larger amount of flue gas can be processed for the faster solid velocity and this decreases the time taken to recover 1 ton CO_2 and also total work.

In CM, the average total work is larger than that for BF and FF beds. This larger total work is resulted from the heat required to HTF. The sensible and reaction energy has to be supplied to HTF without the sensible heat exchange in BF and FF beds. Therefore, final state of exit HTF from absorber has an impact on the process performance. The small amount of circulating HTF increases the exit temperature of HTF; however, the heat content is decreased.

The effects of relative variation of selected parameters on the CO_2 avoided costs are shown in Fig. 4. Optimal conditions of process are different to those in Fig. 3 for several parameters because capital expenditures are included.

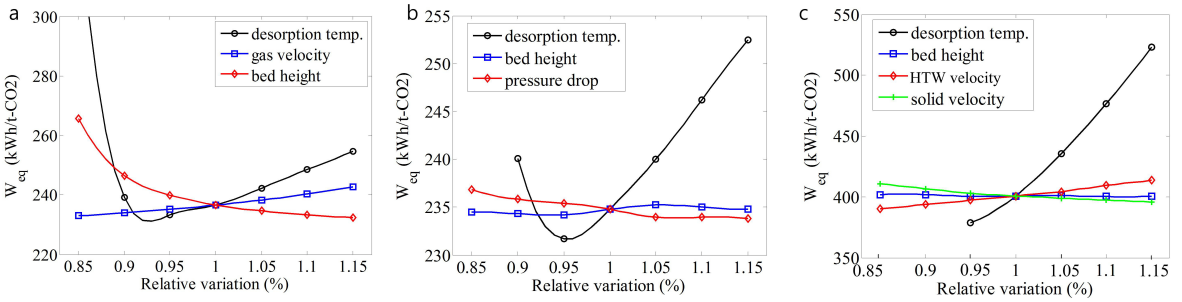

Fig. 3. Effects of perturbation of selected operating condition around their nominal values on the total work, (a): BFB, (b): FFB, and (c): CMB. (CO_2 recovery is constrained at 85 % for all cases).

Despite of this disparity, the trends of effects of parameter change are similar. In BFB, the optimal desorption temperature from the point of the total work and CO_2 avoided cost are different. For latter case, as desorption temperature increases, smaller solid circulation rate and solid hold-up are sufficient to achieve 85 % CO_2 recovery. Smaller solid circulation rate decreases the equipment cost of solid heat exchange due to smaller heat exchange surface area and smaller solid hold-up decreases the inventory cost; these lowers the capital expenditure for relative variation of 0 %. This is also held in FFB where the optimal desorption temperature is different as was in BFB because of same reason.

In CM, the average CO_2 avoided cost is comparative or less than that for BF and FF beds. This is mainly contributed by the absence of solid heat exchanger. Because of high solid circulation rate, the required heat surface area is large to exchange sensible energy such that the estimated surface area is 48,175 m^2 and 56,617 m^2 for BF and FF beds, respectively, and the equipment cost of heat exchanger takes part of 30-50 % of total PFC. Although the reactor dimension is larger for CM to process same flow rate than BF and FF beds due to the restriction of gas velocity under fluidization condition, the absence of heat exchanger lowers the total capital cost. For CM, the heat plate exchanger is embedded in absorber and regenerator as depicted in our previous work [7], which was reflected during cost estimation.

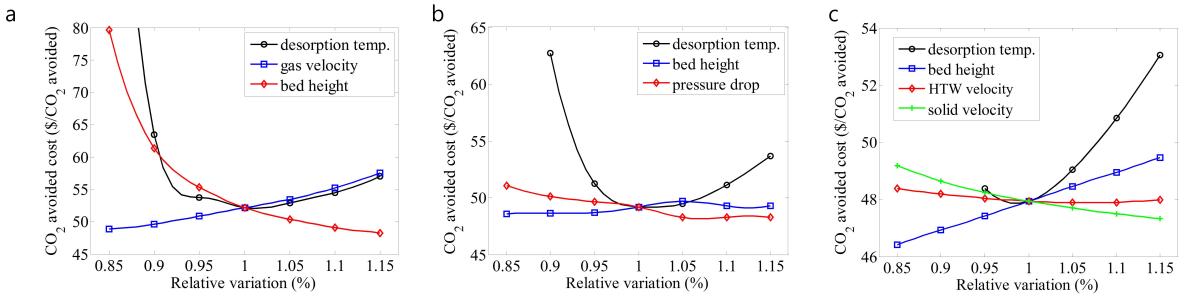

Fig. 4. Effects of perturbation of selected operating condition around their nominal values on the CO2 avoided cost, (a): BFB, (b): FFB, and (c): CMB. (CO_2 recovery is constrained at 85 % for all cases).

4.2. Process evaluation

In Fig. 4, the optimal conditions of operating parameters were chosen and at this condition we evaluated the process performance of BF, FF, and CM beds in Table 4. As a result of comparative study, although CMB has the highest total work because of heating energy to HTF, its PFC and CO_2 avoided cost is lowest due to the absence of external solid heat exchanger. The preliminary capital cost estimate (±30% confidence) has been conducted, however, its relative cost values are more accurate. In addition, in CMB, control of residence time of amine sorbent is much facile than BF and FF beds because the solid and gas velocity are independent variables, which may be a strong advantage to minimize the hydration of amine sorbent by regulating the residence time.

Table 4. Process performance comparison of BF, FF, and CM beds in terms of major parameters when the operating conditions are optimally chosen.

	Solid circulation rate (ton/hr)	W_{total} (kWh/tCO_2)	PFC (Million-$)	CO_2 avoided cost ($/cost avoided)
BFB	3,468	230.75	219.0	47.36
FFB	4,074	234.51	243.1	48.61
CMB	4,149	400.26	135.7	45.77

According to the avoided cost data, CMB is an optimal separation process using amine sorbent; however, its gap with other processes is subtle to declare an optimal process. From the comparison analysis, a research direction of process development can be summarized as following: in fluidization bed processes, an innovative design of solid heat exchanger would be the key for utilization of fluidization process as a gas-solid contactor. The regulation of residence time of amine sorbent to minimize the hydration of sorbent is also important design point; in CMB, minimization of the total work is required by improving internal HTW-gas-solid heat exchange mechanism such as a heat integrated MBA (moving bed adsorption) process shown in the study by Kim et al. [7]. In order to make use of CMB, a particle synthesis method that increase AI of amine sorbent is crucial while maintaining the same performance even particle size grows to the magnitude of mm.

5. Conclusion

The performance of solid sorbent-based carbon capture process using PEI impregnated silica sorbent is evaluated. Selection of proper gas-solid contactor may significantly affect to the process economy along with sorbents performances. Thus, bubbling, fast fluidization and counter-current moving bed are compared to determine an optimal process using amine sorbent. The mathematical modeling of each process was conducted and performances of three models are compared in terms of several economy indicators.

CMB has a higher total work than BF and FF beds because of large heating energy to HTF without self-heat exchange between solid-solid heat exchanger as in BF and FF. However, due to embedded parallel-aligned plate heat exchanger in absorber, the absence of external solid-solid heat exchanger lowers the capital expenditure and so does the CO_2 avoided cost that is calculated to be slightly lower than other processes.

This work suggested that using fluidization or moving beds shows a more lowered CO_2 avoided cost than conventional amine scrubbing process ($61/ton CO_2 [13]). Moreover, the process economy of solid sorbent-based processes may be improved by lowering the equipment cost of solid-solid heat exchanger; enhancing HTF-gas-solid heat exchanger network in CMB; facile regulation method of solid residence time in fluidization processes to avoid sorbent hydration; synthesizing a high AI sorbent of large particle on the order of mm; partly using alternative heat exchange mechanism devised in multi-stage fluidized bed process [8] or heat integrated moving bed process [7].

Acknowledgements

This work was supported by a grand from the Korea CCS R&D Center (KCRC), which is funded by the Korean government (Ministry of Education, Science and Technology, No. 2012-0008886).

References

[1] NETL D. DOE/NETL Carbon Dioxide Capture and Storage RD&D Roadmap. 2011;:1–78.
[2] IEA. Technology Roadmap. 2013;1–63.
[3] Haszeldine RS. Carbon capture and storage: how green can black be? Science. 2009;325:1647–1652.
[4] Charitos A, Hawthorne C, Bidwe AR, Sivalingam S, Schuster A, Spliethoff H, Scheffknecht G. Parametric investigation of the calcium looping process for CO2 capture in a 10kWth dual fluidized bed. International Journal of greenhouse Gas Control. Elsevier Ltd; 2010;4:776–784.
[5] Rodríguez N, Alonso M, Abanades JC. Experimental investigation of a circulating fluidized-bed reactor to capture CO2 with CaO. AIChE Journal. 2010;57:1356–1366.

[6] Modekurti S, Bhattacharyya D, Zitney SE. Dynamic Modeling and Control Studies of a Two-Stage Bubbling Fluidized Bed Adsorber-Reactor for Solid–Sorbent CO 2Capture. Industrial & Engineering Chemistry Research. 2013;52:10250–10260.

[7. Kim K, Son Y, Lee WB, Lee KS. Moving bed adsorption process with internal heat integration for carbon dioxide capture. International Journal of greenhouse Gas Control [Internet]. Elsevier Ltd; 2013;17:13–24..

[8] Kim K, Kim D, Park Y-K, Lee KS. A solid sorbent-based multi-stage fluidized bed process with inter-stage heat integration as an energy efficient carbon capture process. International Journal of greenhouse Gas Control. Elsevier Ltd; 2014;26:135–146.

[9] Veneman R, Kamphuis H, Brilman DWF. Post-Combustion CO. Energy Procedia. Elsevier B.V; 2013;37:2100–2108.

[10] Ströhle J, Junk M, Kremer J, Galloy A, Epple B. Carbonate looping experiments in a 1MWth pilot plant and model validation. Fuel. Elsevier Ltd; 2014;127:13–22.

[11] Bouillon P-A, Hennes S, Mahieux C. ECO2: Post-combustion or Oxyfuel-A comparison between coal power plants with integrated CO_2 capture. Energy Procedia. 2009;1:4015–4022.

[12] Knudsen JN, Jensena JN, Vilhelmsen P-J, Biede O. Experience with CO_2 capture from coal flue gas in pilot-scale: Testing of different amine solvents. Energy Procedia. 2009;1:783–790.

[13] Zhao M, Minett AI, Harris AT. A review of techno-economic models for the retrofitting of conventional pulverised-coal power plants for post-combustion capture (PCC) of CO2. Energy Environ. Sci. 2012;6:25–40.

[14] Wang Q, Luo J, Zhong Z, Borgna A. CO_2 capture by solid adsorbents and their applications: current status and new trends. Energy Environ. Sci. 2011;42:42–55.

[15] Samanta A, Zhao A, Shin, Sarkar P, Gupta R. Post-Combustion CO_2 Capture Using Solid Sorbents: A Review. Industrial & Engineering Chemistry Research. 2012;51:1438–1463.

[16] Sayari A, Belmabkhout Y, Serna-Guerrero R. Chemical Engineering Journal. Chemical Engineering Journal. Elsevier B.V; 2011;171:760–774.

[17] Hoffman JS, Hammache S, Gray ML, Fauth DJ, Pennline HW. Fuel Processing Technology. Fuel Processing Technology. Elsevier B.V; 2014;126:173–187.

[18] Mebane DS, Kress JD, Storlie CB, Fauth DJ, Gray ML, Li K. Transport, Zwitterions, and the Role of Water for CO 2Adsorption in Mesoporous Silica-Supported Amine Sorbents. The Journal of Physical Chemistry C. 2013;117:26617–26627.

[19] Mebane DS, Fauth DJ, Gray ML. A Rigorous Yet Scalable Kinetic Model for the Uptake of CO_2 by Silica-Supported, PEI-Impregnated Sorbents. 2011;:1–15.

[20. Ebner AD, Gray ML, Chisholm NG, Black QT, Mumford DD, Nicholson MA, Ritter JA. Suitability of a Solid Amine Sorbent for CO 2Capture by Pressure Swing Adsorption. Industrial & Engineering Chemistry Research. 2011;50:5634–5641.

[21] Bohart GS, Adams EQ. Some aspects of the behavior of charcoal with respect to chlorine. Journal of the American Chemical Society. 1920;42:523–544.

[22] Monazam ER, Spenik J, Shadle LJ. Fluid bed adsorption of carbon dioxide on immobilized polyethylenimine (PEI): kinetic analysis and breakthrough behavior. Chemical Engineering Journal. 2013;223:795–805.

[23] Kunii D, Levenspiel O. Fluidization Engineering. Stoneham, MA (US): Butterworth Publishers; 1991.

[24] Kim K, Lee S, Ryu JH, Lee KS, Lee WB. An improved CO_2 adsorption efficiency for the zeolites impregnated with the amino group: A molcular simulation approach. International Journal of greenhouse Gas Control. Elsevier Ltd; 2013;19:350–357.

[25] Towler G, Sinnott RK. Chemical Engineering Design, Second Ed.: Principle, Practice and Economics of Plant and Process Design. Butterworth-Heinemann; 2012.

[26] Rao AB, Rubin ES, Berkenpas MB. An integrated modeling framework for carbon management technologies. Department of Engineering and Public Policy, Pittsburgh, PA; 2004. pp. 15213–13890.